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Economic Analysis in Series-Distillation Desalination

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Abstract

The ability to produce potable water economically is the primary purpose of seawater desalination research. Reverse osmosis (RO) and multi-stage flash (MSF) cost more than potable water produced from fresh water resources. Therefore, this research investigates a high-efficiency mechanical vapor-compression distillation system that employs an improved water flow arrangement. The incoming salt concentration was 0.15% salt for brackish water and 3.5% salt for seawater, whereas the outgoing salt concentration was 1.5% and 7%, respectively. Distillation was performed at 439 K and 722 kPa for both brackish water feed and seawater feed. Water costs of the various conditions were calculated for brackish water and seawater feeds using optimum conditions considered as 25 and 20 stages, respectively. For brackish water at a temperature difference of 0.96 K, the energy requirement is 2.0 kWh/m^3. At this condition, the estimated water cost is \$0.39/m^3 achieved with 10,000,000 gal/day distillate, 30-year bond, 5% interest rate, and \$0.05/kWh electricity. For seawater at a temperature difference of 0.44 K, the energy requirement is 3.97 kWh/m^3 and the estimated water cost is \$0.61/m^3. Greater efficiency of the vapor compression system is achieved by connecting multiple evaporators in series, rather than the traditional parallel arrangement. The efficiency results from the gradual increase of salinity in each stage of the series arrangement in comparison to parallel. Calculations using various temperature differences between boiling brine and condensing steam show the series arrangement has the greatest improvement at lower temperature differences.

Keywords: desalination, dropwise condensation, mechanical-vapor compression

1. Pendahuluan

The shortage of fresh water is important because it continuously increases and adversely effects many countries. Water shortages involve more than 80 countries and 40% of the world population [Sharma *et al.*, 2006). There are 1.1 billion people without adequate drinking water. Based on forecasts for 2020, over 60% of humanity will be exposed to water shortages. Currently, about 15,000 desalination units are operating worldwide (Riverol and Pilipovik, 2008). To make desalination technology more attractive, there is a need to lower costs.

According to Aly and El-Fiqi (2003), vapor-compression (VC) distillation is commonly used for small and medium-scale desalination units. There are two methods to compress the vapor: mechanical compressors and steam jets. Lara (2005) states that mechanical vapor compression is very efficient. Unlike other distillation systems, it does not require a large external heating source; however, it requires very skilled operators and has higher maintenance costs compared to thermal vapor compression (Al-Juwayhel *et al.*, 1997).

In Mechanical-Vapor-Compression (MVC) system, an electric motor or diesel engine compresses the water vapor thus raising its pressure and saturation temperature. This temperature difference is essential for the evaporation process. Capacities and pressure ratios of the vapor compressors play major roles in MVC systems. Small inter-effect temperature differences minimize the mechanical energy input required to drive the compressor. One design employed a centrifugal compressor that has a compression ratio of about 1.6 (Senatore, 1981).

Lara (2005) states that increasing the seawater concentration elevates its boiling temperature and reduces its vapor pressure. His evaporator uses a mechanical vapor compression technology for the separation. This section briefly elaborates on pressure and boiling point elevation used in this research that affect evaporator performance.

Lucas and Murat (1985) state that the seawater desalination process has low energy consumption, an important characteristic, which can reach 9 kWh per m^3 of product water. It is located at nuclear power plant in

Flamanville, France, and produces 1,500 m^3/day of high-quality product water.

In a rather complicated study, Lucas and Tabourier (1985) showed that this optimization is reached for a compression ratio that gives an overall temperature difference of 13°C between the first and fourth effects. This agrees with the initial choice of the process factors, such as limited temperature differences between each cell or between the tube bundles and the evaporator cells. The water vapor pressure of seawater and its concentrates have been measured from 373 K to 353 K by Emerson and Jamieson (1967). The results of their measurements are close to the analytical method described by The National Engineering Laboratory of England. The vapor pressure p_o of pure water at a measured temperature can be obtained from steam tables or it can be calculated as follows:

$$\log_{10} p_o = a + \frac{b}{z} + \frac{cx}{z}\left(10^{dx^2} - 1\right) + e10^{fy^{1.25}} \quad (1)$$

where

p_o = pure water vapor pressure (10^5 N/m^2)
$x = z^2 - g$
$y = 344.11 - t$
$z = t + 273.16$
t = measured temperature (°C)
$a = 5.432368$
$b = -2.0051 \times 10^{3}$
$c = 1.3869 \times 10^{-4}$
$d = 1.1965 \times 10^{-11}$
$e = -4.4000 \times 10^{-3}$
$f = -5.7148 \times 10^{-3}$
$g = 2.9370 \times 10^{5}$

For a wet compressor, the isentropic compressor work is evaluated by Lara (2005) as

$$W = \frac{(1+x)\hat{H}_2^{vap} - (\hat{H}_1^{vap} + x\hat{H}_1^{liq})}{\eta_c} \quad (3)$$

where

$\hat{H}_2^{vap}$ = vapor enthalpy at compressor exit (2) (J/kg)
$\hat{H}_1^{vap}$ = vapor enthalpy at compressor inlet (1) (J/kg)
$\hat{H}_1^{liq}$ = liquid enthalpy at compressor inlet (J/kg)
η_c = compressor efficiency = 0.85
x = the amount of injection water that evaporates in the compressor

$$x = \frac{S_1^{vap} - S_2^{vap}}{S_2^{vap} - S_1^{liq}} \quad (4)$$

where

S_1^{liq} = entropy of liquid water at compressor inlet (J/(kg·K))
S_1^{vap} = entropy of steam at compressor inlet (J/(kg·K))
S_2^{vap} = entropy of steam at compressor exit (J/(kg·K))

The boiling point elevation corresponding to each measured value of vapor temperature is plotted against the salinity in Figure 1. The research is performed in two stages: (1)

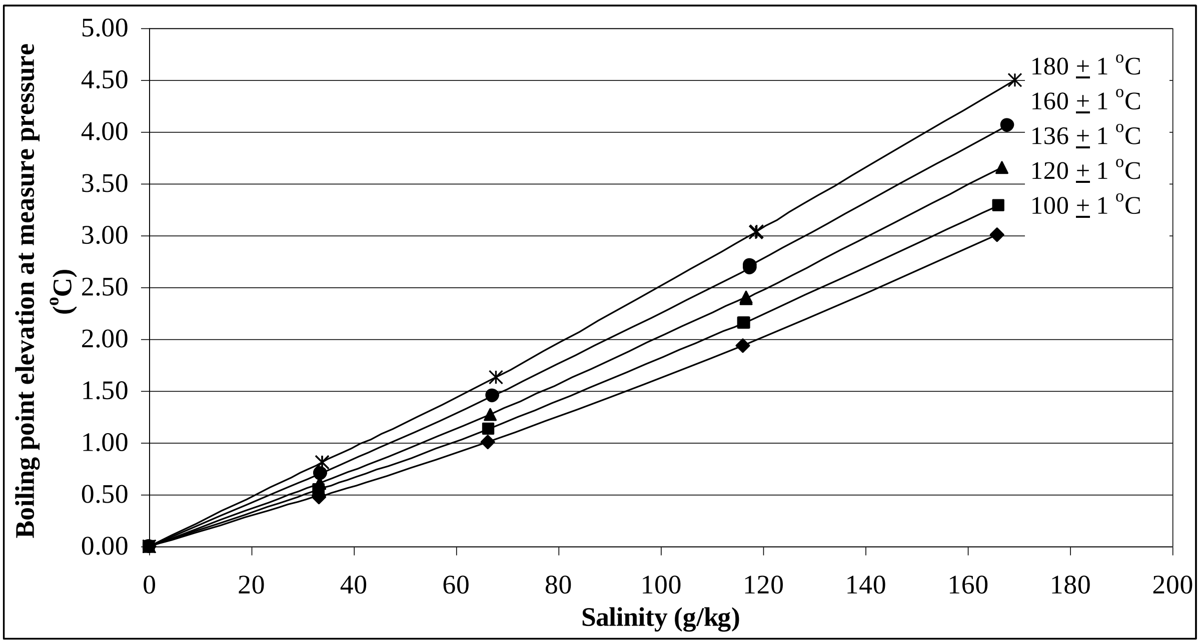

Figure 1. Boiling point elevation and salinity at various temperatures.

comparison of series and parallel flow arrangements and (2) economic analysis.

The specific objective for this project is to find the optimal cost of potable water ($/m^3) produced by series-distillation desalination.

2. Methodology

Degassed seawater supplied to the evaporator trains is passed through the sensible heat exchanger shown in Figures 2 and 3.

The seawater salinity is 35 g salt/kg solution. Then the seawater is fed upflow into the

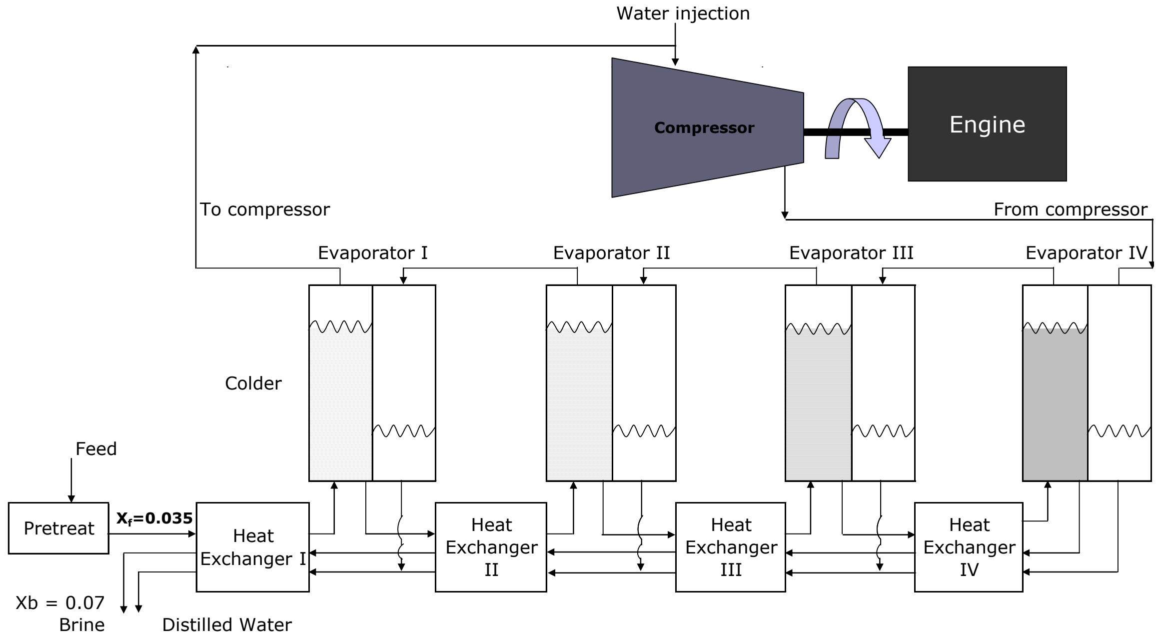

Figure 2. Series vapor-compression desalination.

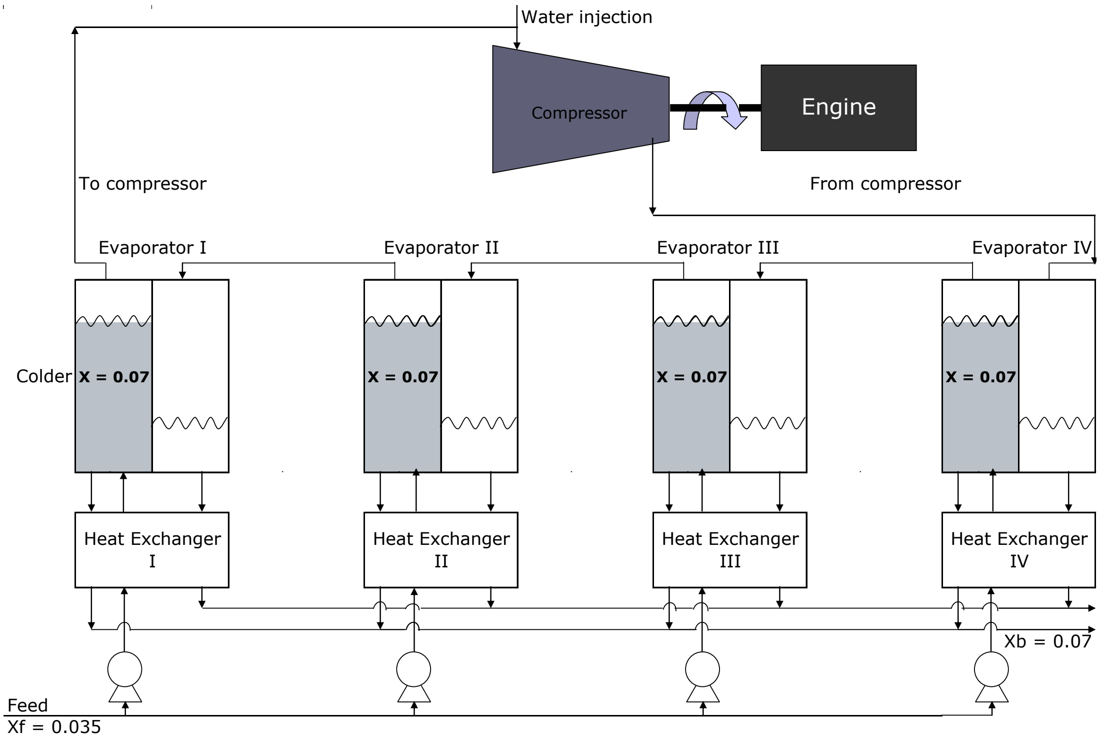

Figure 3. Parallel vapor-compression desalination.

latent heat exchangers. Saturated steam is supplied from various temperature differences. Three trade-off cases will be investigated with ΔT 3.333 K, 2.222 K, and 1.111 K. In all cases, the brine salt concentration is 70 g salt/kg brine. The maximum pressure on the steam side is limited to 120 psig to ensure dropwise condensation. The design is performed with series and parallel evaporators to determine the effect on energy efficiency, as summarized in Table 1. The results obtained are useful to design systems and to evaluate the economic perspectives of this technology.

Table 1. MVC base system.

Design parameters	Unit	Value	
Feed water salinity	g/kg	1.5	35
Brine salinity	g/kg	15	70
Plant capacity	m^3/s	0.4381	
Feed water temperature	K	294	
Steam pressure	kPa	418; 427; 722	
ΔT in latent heat exchanger	K	0.19; 0.39; 0.56; 1.11; 1.67; 2.21	
Interest rate	%	5; 10; 15; 20	
Electricity	\$/kWh	0.05; 0.10; 0.15	
Plant lifetime	Year	30	

Number of stages used will be determined based on the data.

To begin the economic analysis, a hypothetical base system is developed that employs MVC to desalt feed water. The feed waters are brackish and seawater, with salinities 1.5 g salt/kg solution and 35 g salt/kg solution, respectively. Based on the salinities, a recovery rate of the MVC unit can be determined. The recovery rate (*RR*) is determined by

$$RR = \frac{f_P}{f_f} \times 100\,\% \qquad (5)$$

where

f_P = the product water flow rate (m^3/s)
f_f = the feed water flow rate (m^3/s)

The distillate production capacity in the economic analysis is 0.4381 m^3/s. Single-stage vapor-compression desalination system is used as base in the economic evaluation. The amount of brackish water feed required to supply the distillate flow rate is calculated by the corresponding mass balance. The feed temperature is 294 K. The evaporator is constructed with naval brass with a coating that promotes dropwise condensation. Heat transfer coefficients of the evaporator for each condition come from measurements. The heat duty for the latent heat exchanger is calculated by

$$Q = mL \qquad (6)$$

where,

Q = the amount of energy required to change the water phase (J/s)
m = the mass of the distillate (kg/s)
L = the specific latent heat for distillate (J/kg)

The heat exchanger area is given by U. S. Environmental Protection Agency (1996) as follows.

$$A = \frac{Q}{U\Delta T} \qquad (7)$$

where,

A = area of heat transfer surface (m^2)
Q = amount of heat transferred to distillate from evaporator (J/s)
U = overall heat transfer coefficient ($J/(s{\cdot}m^2{\cdot}K)$)
ΔT = temperature difference in latent heat exchanger (K)

The total capital investment is calculated by selecting the overall temperature difference. The cost model for the VC desalination system consists of both operating costs and capital costs associated with purchased equipment and installation.

3. Results and Discussion

3.1 Energy Comparison of Series and Parallel Flow Arrangements

Degassed seawater with 3.5% salt supplied to the evaporator train is connected in series and parallel to satisfy individual evaporator temperature needs. The feed flow rate is 295 kg/s for series and parallel flows.

Four evaporator stages are assumed, each unit with ΔT = 3.333 K, 2.222 K, 1.111 K, and 7% brine. At lower Δ*T*, the compressor shaft work requirements are lower. The work requirements of series and parallel mechanical vapor-compression desalination were compared to determine the relative efficiency.

The seawater passed through the evaporator trains is shown in Figure 1. The temperature and pressure were calculated at the inlet and outlet of each evaporator. From these state values, enthalpy and entropy of water were determined using steam tables. The compressor work was calculated for each Δ*T*.

The seawater entering the evaporators is also connected in parallel (Figure 2). The energy analysis was repeated for parallel desalination.

The results in Table 2 show that the series configuration is more efficient than the parallel configuration. The efficiency improvement is larger for small Δ*T* because the boiling point elevation of the salt water is a larger portion of the overall Δ*T*.

Based on these results, only the series configuration will be used to calculate the cost of water produced from brackish and seawater. These values show that pressure differences of compressor for series desalination are lower than those of parallel desalination. Vapor formed in the first latent heat exchanger goes to the compressor where its pressure and saturation temperature are raised.

The vapor pressure of series desalination is higher than that of parallel desalination because of its lower salinity. The higher vapor pressure in the first latent heat exchanger results in the lower pressure difference of the compressor.

Power consumption of the compressor, and therefore the efficiency of the process, is proportional to this pressure difference. By lowering the pressure difference, it is possible to decrease the energy consumption of the process.

Table 2. Percent reduction in compressor power consumption for series desalination compared to parallel desalination.

Δ*T* (K)	Reduction in Power Consumption (%)
1.111	15.21
2.222	10.80
3.333	8.37

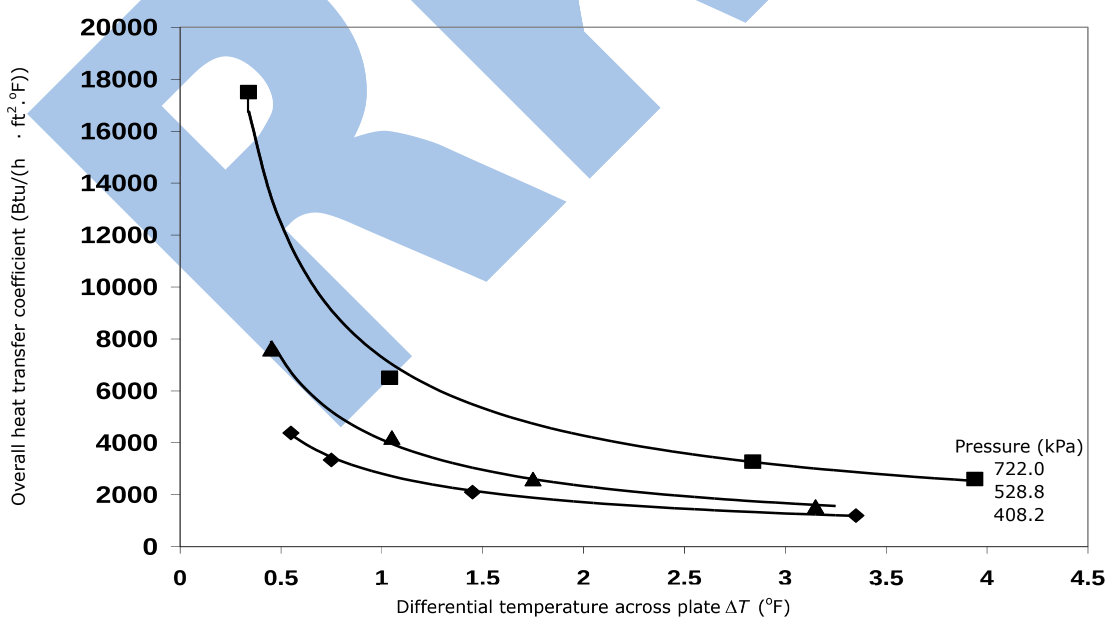

Figure 4. Measured heat transfer coefficients for dropwise condensation of pressurized steam.

Both the values reported in Table 2 and values from the study by Tleimat (1995) show energy savings from the series arrangement. However, the two studies cannot be compared because the equipment was different.

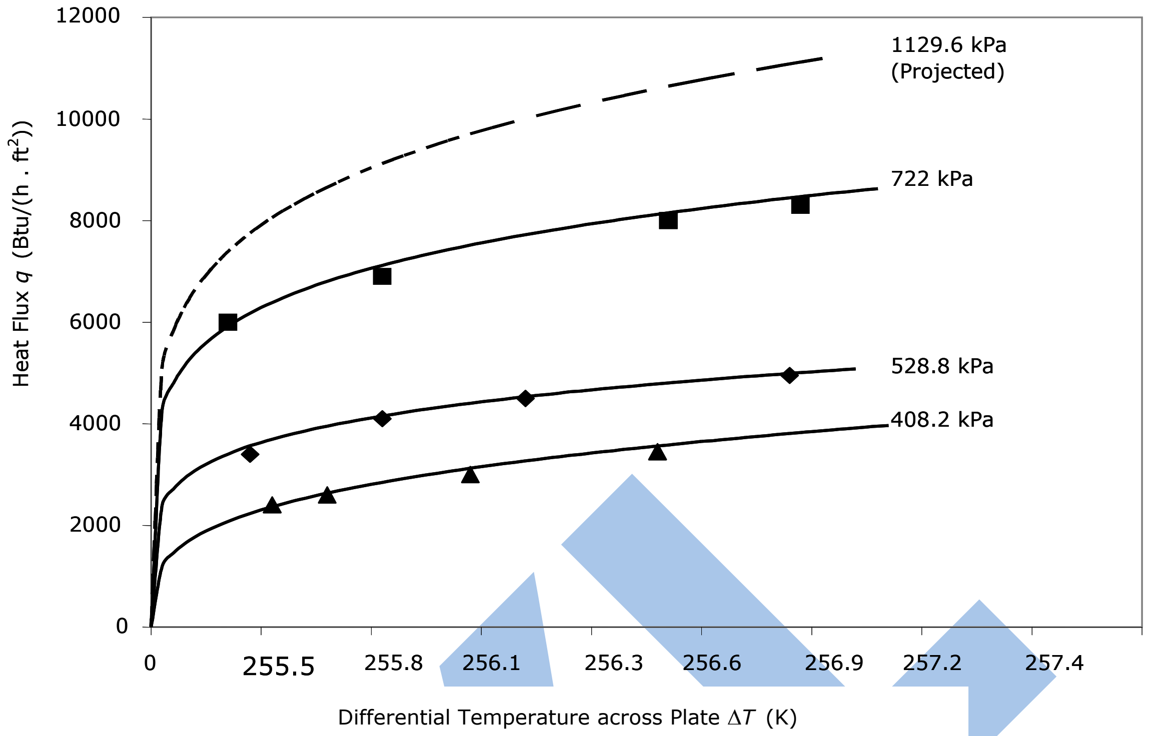

Figure 5. Effect of temperature difference on latent heat exchanger heat flux.

Both the values reported in Table 2 and values from the study by Tleimat (1995) show energy savings from the series arrangement. However, the two studies cannot be compared because the equipment was different.

The study by Tleimat compared actual energy consumption of series multi-effect vapor compression distillation to that of single-effect distillation. Based on Tleimat's study, actual savings in the energy consumption are higher than the values in Table 2 and depend on the number of effects. In contrast, the research done here compares series to parallel multi-effect vapor-compression distillations.

3. 2 Economic Analysis

Lara (2005) shows that the economic latent heat exchanger ΔT recommended for the United States and the Middle East are 1.111 K and 3.333 K, respectively (Lara, 2005). The ΔT has a large effect on the compressor work.

At lower ΔT, the compressor work needed to increase the temperature is lower. However, if the ΔT is too low, a larger heat exchanger is needed, which is not economical. Heat transfer coefficients of the evaporator for each condition are shown in Figure 4.

Table 3. Required areas of heat exchangers at various pressures.

Description	Brackish water feed	Seawater feed
Latent heat exchanger temperature difference (K)	1.111	0.389
Number of stages	25	20
Inlet distillate velocity in sensible heat exchanger (m/s)	14	14
Distillate pressure drop in sensible heat exchanger (kPa/m)	576	561
Heat transfer areas at 104.7 psia (722 kPa)		
-Latent heat exchanger (m^2)	36,100	54,000
-Sensible heat exchanger (m^2)	13,300	20,500
-Total areas (m^2)	49,400	63,500
Heat transfer areas at 76.7 psia (528.8 kPa)		
-Latent heat exchanger (m^2)	62,500	75,500
-Sensible heat exchanger (m^2)	11,400	18,700
-Total areas (m^2)	73,900	94,200
Heat transfer areas at 59.2 psia (408.2 kPa)		
-Latent heat exchanger (m^2)	86,400	117,000
-Sensible heat exchanger (m^2)	9,900	17,000
-Total areas (m^2)	96,300	134,000

The heat flux is calculated by multiplying the heat transfer coefficient by the temperature difference. The figure shows that above $\Delta T \approx$

256.1 K, the heat flux is virtually independent of temperature difference (Figure 5).

Nonetheless, to find the economic optimum, the explored ΔT will range from 0.19 to 2.21 K. Figure 5 shows a strong benefit from operating at higher pressures, so economic calculations will focus on a selected pressure of 722 kPa. Based on the temperature differences and pressures used, the areas of both latent and sensible heat exchangers can be calculated.

The sensible heat exchanger is a key component of the desalination system. To improve heat transfer, a microchannel design is employed. It consists of three plates.

The distillate from the latent heat exchanger enters into the first plate. The feed water or brine from previous latent heat exchanger enters into the second plate. The brine from the last latent heat exchanger enters into the third plate. Inlet distillate velocity was determined based on the same pressure drop <52.5 kPa/m for each microchannel heat exchanger.

Table 3 shows the effect of operating pressure on total heat exchanger area (latent plus sensible heat exchangers). The total area strongly depends on the latent heat exchanger area. All calculation results showed that the higher the pressure in the latent heat exchanger, the smaller the total areas; therefore, to minimize cost, the water cost calculation only focuses on the highest pressure (722 kPa). The sensible heat exchanger can be produced using titanium-coated naval brass. The titanium coat provides a tough surface that resists abrasion and reduces fouling whereas the naval brass core provides good heat transfer. With a wall thickness of 1.5 mm (0.059 inches) as a standard sold in the market, the prototype is counter-current microchannel heat exchanger with single-passage microchannels. All the channels have the same length, making it

Table 4. Water cost for brackish water feed at three electricity costs.

Cost for brackish water feed	Electricity cost[a] $0.05/kWh		Electricity cost[b] $0.1/kWh		Electricity cost[b] $0.15/kWh	
	$\$/m^3$	$/yr	$\$/m^3$	$/yr	$\$/m^3$	$/yr
Electricity	0.110	1,532,969	0.135	2,094,736	0.202	2,803,334
Labor[c]	0.036	500,000	0.036	500,000	0.036	500,000
Bond (5%, 30 years)	0.131	1,810,891	0.170	2,320,413	0.170	2,353,734
Maintenance (0.04 x FCI)	0.081	1,113,514	0.105	1,426,818	0.105	1,447,307
Insurance (0.005 x FCI)	0.010	139,189	0.013	178,352	0.013	180,913
Total	0.368	5,096,563	0.459	6,520,319	0.526	7,285,289
Brine injection well	0.041	559,488	0.041	559,488	0.041	559,488
Total	0.409	5,656,051	0.500	7,079,807	0.567	7,844,777

[a]Latent heat exchanger temperature difference is 1.111 K
[b]Latent heat exchanger temperature difference is 0.556 K
[c]Source: RosTek Associates, Inc., *Desalting Handbook for Planners*, 3rd ed.

Table 5. Water cost for seawater feed at three electricity costs.

Cost for seawater feed	Electricity cost[a] $0.05/kWh		Electricity cost[a] $0.1/kWh		Electricity cost[b] $0.15/kWh	
	$\$/m^3$	$/yr	$\$/m^3$	$/yr	$\$/m^3$	$/yr
Electricity	0.194	2,684,199	0.389	5,368,399	0.536	7,411,639
Labor[c]	0.036	500,000	0.036	500,000	0.036	500,000
Bond (5%, 30 years)	0.193	2,672,029	0.193	2,672,029	0.215	2,964,897
Maintenance (0.04 x FCI)	0.119	1,643,025	0.119	1,643,025	0.132	1,823,110
Insurance (0.005 x FCI)	0.015	205,378	0.015	205,378	0.017	227,889
Total	0.557	7,704,631	0.752	10,388,831	0.936	12,927,535
Ion Exchange Unit	0.055	760,072	0.055	760,072	0.055	760,072
Total	0.612	8,464,704	0.807	11,148,903	0.991	13,687,607

[a]Latent heat exchanger temperature difference is 0.389 K
[b]Latent heat exchanger temperature difference is 0.189 K
[c]Source: RosTek Associates, Inc., *Desalting Handbook for Planners*, 3rd ed.

possible to minimize the variance of the residence time distribution. The costs of sensible and latent heat exchangers are \$1.96/m^2 and \$0.81/m^2, respectively. With low-cost manufacturing as investigated by Lara (2005), it appears that less expensive heat exchanger can be made using naval brass 464.

This research showed that when brackish water is used, the minimum water cost can be achieved by using 25 stages. When seawater is used, the system requires 20 stages to get minimum water cost. Based on the various stages, the cost of water is calculated.

Tables 4 and 5 show the cost of desalinated water at three electricity costs: \$0.05/kWh, \$0.1/kWh, and \$0.015/kWh. In these tables, very small temperature differences 0.189 K – 1.111 K were employed. For all electricity costs, potable water from brackish water feed is less expensive (\$0.41/m^3) than seawater feed (\$0.61/m^3). The water costs are achieved at 5% interest rate and 722 kPa and as much as \$0.39/m^3 for brackish water feed.

4. Conclusion

Series desalination requires less work than parallel desalination. The percent reduction in power consumption increases with decreasing temperature difference in the evaporator. Compared to parallel desalination, series vapor-compression desalination reduces power consumption by 8.37 – 15.21%. This savings is achieved because much of the water is removed at a lower salt concentration, which has a higher vapor pressure and requires less compression ratio. The improvement is more pronounced at small temperature differences in the evaporator (1.111 to 2.222 K).

Optimum conditions for mechanical vapor-compression desalination were determined. The latent heat exchanger employs dropwise condensation, which allows economical operation with very low temperature difference in the evaporator, which makes the system more efficient. The water costs at 722 kPa and \$0.05/kWh are \$0.39/m^3 and \$0.61/m^3 for brackish water feed and seawater feed, respectively.

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